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Dynamic Modeling and Simulation of MSF Desalination Plants

Awad S. Bodalal*, Sayed A. Abdul_Mounem, Hamid S. Salama

Mechanical Engineering Department, Faculty of Engineering, Garyounis University, Benghazi-Libya

Abstract

This paper describes the development of a mathematical model to predict the performance of multi-stage flash (MSF) plant systems under transient conditions. The model developed is based on coupling the dynamic equations of mass, energy and momentum. These equations describe the dynamic behavior of brine and product streams within the flashing stages, the effect of salinity and temperature variation on the specific heat, boiling point, enthalpy and density is accounted for in this model. The model which consists of a set of differential and algebraic equations (describing the dynamic behavior of each stage in terms of some key physical parameters) are solved by using the fifth order Runge-Kutta method. The proposed model was validated by using data from previous theoretical studies as well as actual data obtained from an operating MSF plant. The results obtained are useful for dynamic parametric studies and for the prediction of the performance of a given plant under a wide range of possible transient conditions.

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Keywords. Desalination; MSF; Dynamic Modeling; Simulatio

Nomenclature

A_{BH}	Brine heater tube section area
$Cp_{b(i)}$	Specific heat of flashing brine stream exiting the i^{th} stage
$Cp_{ba(i)}$	Specific heat of circulating brine exiting the i^{th} stage
$Cp_{bd(i)}$	Specific heat of blow down
$Cp_{bg(i)}$	Specific heat of vapor flashing from brine pool in the i^{th} stage
Cp_{BH}	Specific heat of brine in brine heater
$Cp_{C(i)}$	Specific heat of liquid condensing from the tubes in the i^{th} stage
$Cp_{gv(i)}$	Specific heat of vapor in the i^{th} stage
Cp_K	Specific heat of make up
$Cp_{pg(i)}$	Specific heat of vapor flashing from distillate tray in the i^{th} stage
$Cp_{pr(i)}$	Specific heat of the product in the i^{th} stage
$Cp_{V(i)}$	Specific heat of vapor in the i^{th} stage
$h_{bg(i)}$	Enthalpy of vapor flashing from the brine pool in the i^{th} stage
$h_{C(i)}$	Enthalpy of condensing from the cooling tubes in the i^{th} stage
h_L	Enthalpy of condensate in the brine heater
$h_{pg(i)}$	Enthalpy of vapor flashing from distillate tray in the i^{th} stage
h_S	Enthalpy of heating steam
i	Stage number
LT_{BH}	Brine heater tube length
$M_{B(i)}$	Mass hold-up of flashing brine in the i^{th} stage
$m_{b(i)}$	Flashing brine flow rate from the brine pool in the i^{th} stage
m_{bd}	Blow down flow rate
$m_{bg(i)}$	Rate of vaporization from the flashing brine pool in i^{th} stage
M_{BH}	Mass hold-up in the brine heater
$m_{C(i)}$	Rate of condensing from the cooling tubes in the i^{th} stage
$m_{gv(i)}$	Flow rate of vapor due to venting in the i^{th} stage
m_k	Make up flow rate
m_L	Condensate flow rate
$m_{pg(i)}$	Rate of vaporization from the distillate tray in the i^{th} stage
$M_{PR(i)}$	Mass hold-up of distillate in the i^{th} stage
$m_{pr(i)}$	Distillate flow rate from the i^{th} stage
m_S	Steam flow rate
$M_{V(i)}$	Mass hold-up of vapor in the i^{th} stage
n	Last stage
$m_{wa(i)}$	Circulating brine flow rate brine in the i^{th} stage
$M_{WA(i)}$	Mass hold-up for condenser tubes in the i^{th} stage
NT_{BH}	Number of tubes in the brine heater
$T_{B(i)}$	Temperature of the mass hold-up in the flashing brine pool
$T_{b(i)}$	Temperature of flashing brine exiting the i^{th} stage
T_{bd}	Temperature of blow down
$T_{bg(i)}$	Temperature of vapor flashing from the brine pool in the i^{th} stage
$T_{C(i)}$	Temperature of liquid condensing from the tubes in the i^{th} stage
$T_{gv(i)}$	Temperature of vapor in the vapor space in the i^{th} stage

* Corresponding author. *drawadbodalal@hotmail.com*.

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T_L Temperature of condensate
$T_{pg(i)}$ Temperature of vapor flashing from the distillate tray in the i^{th} stage
$T_{PR(i)}$ Temperature of the mass hold-up in the distillate tray
$T_{pr(i)}$ Temperature of distillate exiting the i^{th} stage
T_S Steam temperature
$T_{V(i)}$ Temperature of vapor in the i^{th} stage
$T_{WA(i)}$ Temperature of the mass hold-up in the condenser tubes in the i^{th} stage
$T_{wa(i)}$ Temperature of circulating brine exiting the i^{th} stage
V_{BH} Volume of the brine in the brine heater
$X_{B(i)}$ Salinity of flashing brine in the i^{th} stage
$X_{b(i)}$ Salinity of flashing brine stream exiting the i^{th} stage
X_{bd} Blow down salinity
X_K Make up flow rate salinity
$X_{WA(i)}$ Salinity of circulating brine in the i^{th} stage
$X_{wa(i)}$ Salinity of brine in the condenser tubes exiting the i^{th} stage
Greek
ρ_{BH} Brine density in the brine heater
ΔT Log mean temperature difference

1. Introduction

In arid, water-scarce parts of the world, such as the Arab Gulf area and North Africa, multi-stage flash (MSF) desalination is considered one of the most common techniques that provide a considerable quantity of potable water. Massive amounts of seawater and therefore a similarly large quantity of energy are processed through these large, costly facilities. In that regard, considerable quantities of concentrated brine are disposed of after the desalination treatment. The aforementioned features of MSF plants make topics such as optimization of the operation and minimization of the corresponding environmental impact a main priority [1]. To the aim of addressing all these aspects, mathematical models provide a very useful tool. The steady state modeling of the MSF desalination plant has been the subject of various studies in the past; several publications for the steady-state simulation of desalination systems have been reported in literatures [2-4]. Due to the fact that the MSF desalination process has many input and output variables, it is difficult to model the dynamic behavior of the plant; but to design the instrumentation and control system to have better control on process variables, the knowledge of dynamics is necessary. The first attempt to obtain a dynamical model of an MSF-process was presented in [5]. The first attempt to obtain a dynamical model of an MSF-process was presented by Yokoyama et al. [6]. Viral et al. [7] carried out a dynamic model applying empirical corrections for the evaporation rates. The degrees of freedom based on a dynamic model were used to determine the number of controlled and manipulated variables. Thomas et al. [8] developed a mathematical model and its solution procedure to simulate the dynamic behavior of multi-stage flash desalination plants. The model was used to predict the operating parameters of an actual MSF plant. Theoretical models which simulate the transient behavior of MSF desalination plants under various conditions have been reported by Rimawi et al. [9]. Tarifa and Scenna [10], studied a dynamic simulator for MSF desalination plants, this simulator was developed to study the effects of faults that may affect a MSF system. Other models have also been proposed in [11-15]. In this way, for any process, the development of a dynamic simulator, which is able to simulate a dynamic process, is an interesting goal. Indeed, this dynamic simulator can be used fortraining operating personnel and investigating plant behavior under dynamic situations, which helps to predict the dynamic conditions of the plant in order to study potential operating modes and control behavior. Therefore, this work outlines the development and performance of a simulator able to simulate a dynamic process, specifically aimed to MSF processes at start-up, this simulator was developed to investigate the effects of changing the characteristics of the plant that may affect a MSF system at start up

2. MSF Process Description

MSF desalination is an evaporating and condensing process. The heat energy required for evaporation is supplied by exhaust heat recovery boilers and auxiliary boilers. The energy supplied during evaporation is recovered in the condensation. The MSF unit can be divided into three sections; a heat reject section, a heat recovery section, and a brine-heater section. A schematic diagram for the MSF system with brine circulation is shown in Figure 1. [16]. The recovery and reject sections are made up of a series of stages, each MSF stage has a flash chamber and a condenser; the vapor flashed off in the flash chamber is separated from the condenser by a demister which intercepts brine droplets entrained in the flashing vapor. A distillate trough under the condenser tubes collects the condensate. The sea water from supply pumps enters the MSF plant and flows through the condenser tubes of the reject section stages and gets heated by the heat released due to condensation of the flashing vapor. A part of this stream leaving the reject section is mixed with the sea water to preheat it; a part is added to the flash chamber of the last flash as make-up, and the remainder is rejected to the sea, the recycle brine drawn from the last flash enters the tubes of the recovery section and gradually gets heated by the heat released from the condensation of the flashing vapor in each stage and leaves the recovery section out of the first flash. It enters the brine-heater section on the tube side where it is heated further to the required top brine temperature (TBT) by condensing steam on the shell side. The heated brine then enters the flash chamber of the first stage which is maintained at an appropriate pressure where it flashes. The flashing brine flows from one stage to the next through an orifice which controls the brine level in each flashing chamber. Pressure is gradually decreased in the successive stages as flashing continues. A part of the concentrated brine from the flashing chamber of the last stage is discharged as blow-down and the remainder combined with the make-up flow serves as the recycled brine. The distillate flowing from stage to stage in the distillate tray is taken out as the product from the last stage and chemically treated to adjust the pH and hardness prior to sending it to the storage tanks. There are a number of process variables which need to be set for reasonable operation of an MSF plant. They are:

1. TBT from the brine heater. This directly affects the distillate production and the levels in each flash chamber. There is a maximum allowable value, depending upon the type of scale inhibitors added to the make-up feed.
2. Brine recycles flow. This directly affects the levels in each flash chamber and the steam consumption for a fixed TBT. The higher the flow rate, the lower the flashing efficiency with a reduction in the residence time in the stages and an increasing brine level in the stages.
3. Make-up flow. This makes up for the blow-down flow and the distillate product flow out of the plant. It affects the temperature of the recirculating brine and thus affects the flashing process.
4. Low pressure (LP) steam temperature leaving the spray system. This dictates the heat content of the steam.
5. Sea water feed flow. This governs the fluid velocities in the tubes of the reject section. It affects the heat transfer in the heat reject section.
6. Sea water feed temperature. This directly affects the heat transfer in the reject section. It also affects the temperature of the makeup and thus of the recirculating brine.
7. Brine heater condensate level. This ensures that the heat exchanger tubes are not submerged in condensed steam, since that will adversely affect the heat transfer.
8. Brine level in the last stage. This affects the level of brine in the preceding stages, and helps to avoid drainage of the system.
9. Distillate level in the last stage. This ensures that the distillate does not overflow the distillate tray.

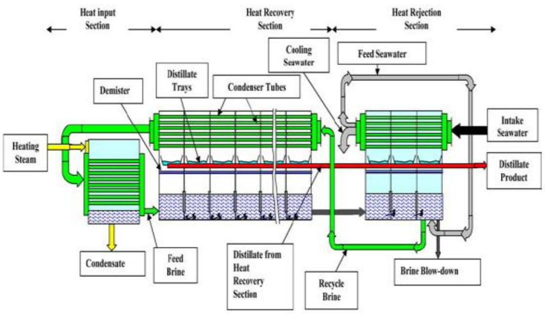

Figure 1. Multi stage flash desalination process [16]

3. Physical Model Description

The MSF process is a flash evaporation process at low pressure (vacuum), where the pressure decreases and the evaporation temperature in accordance decreases from the first to the last stage. From the modeling point of view, it is easier to describe a single flash stage if it is split up into four control volumes, which can be treated separately. The four control volumes are, flash chamber, vapor space, product tray, and tube bundle. Figure 2 shows the graphical representation of the four control volumes of stage. Each fluid stream communicating with the individual stage has four characterizing variables, flow rate, temperature, pressure, and salt concentration. The physical properties, enthalpy, density, and specific heat of the stream, are functions of these variables. The time derivatives included for the mass hold-up, concentration, temperature and specific enthalpy in each of the control volumes of the MSF stages represent the dynamic model. With these derivatives put to zero, the model represents steady-state conditions.

3.1. Mathematical representation of the stage

The MSF desalination stage can be divided into the following four control volumes (CV):

1. The flashing brine pool.
2. The distillate (product) tray.
3. The vapor space.
4. The condenser tubes.

These control volumes are shown in Figure 2. In each control volumes the mass flow rate, temperature, concentration and pressure are considered the fundamental independent variables that characterize the stream. In the present model, the lumped capacitance and uniform state approximations are assumed. Therefore, the state of the flow at CV exit is assumed the same as that for the mass hold-up in the CV. Thus yield;

$T_{b(i)} = T_{B(i)}$, $T_{pr(i)} = T_{PR(i)}$, $T_{wa(i)} = T_{WA(i)}$ and $X_{b(i)} = X_{B(i)}$

Also the following basic assumptions are carried out in the

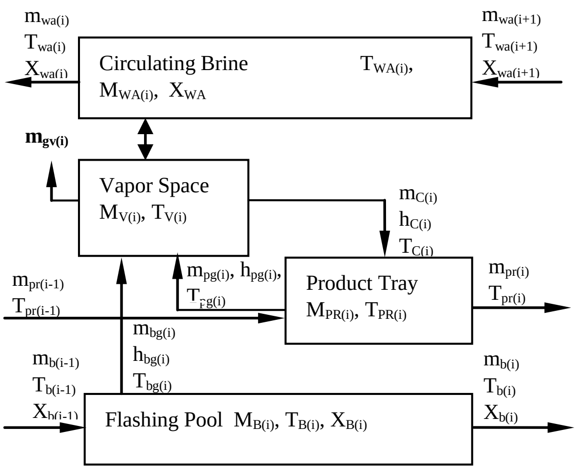

Figure 2. Block diagram of a repeated stage.

present work.

1. Salt is not present in the vapor.
2. Liquid is perfectly mixed on each unit.
3. The heat of mixing due to change in salinity is ignored.
4. Vapor is perfectly mixed on each unit.
5. The vapor is saturated.
6. The oscillations of stage brine level are negligible.
7. Initially at time t = 0, the system is at atmospheric conditions.

in addition to the above-mentioned assumptions, the present model has the following advantages:

1. The pre-heater heat transfer area and the surface area of each flashing chamber are considered in the objective function and the main brine heater transfer area is also considered.
2. The system is well insulated.
3. Dependence of heat capacity, boiling point elevation and latent heat of evaporation on temperature and concentration is considered by rigorous correlations.
4. Dependence of the overall heat transfer coefficient on brine velocity, temperature, and tube diameter is considered.
5. Non-equilibrium allowance is taken into account according to the correlation developed in [3].
6. Hydraulic correlations given in [4]. These equations describe the inter-stage flow rate of the flashing brine.
7. Stage geometric design (length, width and height) is considered.
8. Non-condensable effects are neglected.
9. Maximum number of stages N =21.

3.1.1. Stage Model

The mass conservation equation for the brine in the flashing pool is given by:

$$\frac{dM_{B(i)}}{dt} = m_{b(i-1)} - m_{b(i)} - m_{bg(i)} \quad (1)$$

$$m_{b(i)} = \rho_{b(i)} A_{O(i)} v_{(i)} \quad (2)$$

The balance of the dissolved solids reads

$$M_{B(i)} \frac{dX_{B(i)}}{dt} = m_{b(i-1)} X_{b(i-1)} - m_{b(i)} X_{b(i)} \quad (3)$$

The energy balance for brine pool is given by

$$M_{B(i)} Cp_{b(i)} \frac{dT_{B(i)}}{dt} = m_{b(i)} Cp_{b(i-1)} T_{b(i-1)} - m_{b(i)} Cp_{b(i)} T_{b(i)} - m_{bg(i)} h_{bg(i)} \quad (4)$$

The mass conservation equation for the product in the tray is

$$\frac{dM_{PR(i)}}{dt} = m_{pr(i-1)} + m_{c(i)} - m_{pr(i)} - m_{pg(i)} \quad (5)$$

The energy balance for the product tray is

$$M_{PR(i)} Cp_{pr(i)} \frac{dT_{PR(i)}}{dt} = m_{pr(i-1)} Cp_{pr(i-1)} T_{pr(i-1)} - m_{pg(i)} h_{pg(i)} - m_{pr(i)} Cp_{pr(i)} T_{pr(i)} + m_{c(i)} h_{c(i)} \quad (6)$$

The mass balance for the vapor space produces

$$\frac{dM_{V(i)}}{dt} = m_{bg(i)} + m_{pg(i)} - m_{c(i)} - m_{gv(i)} \quad (7)$$

The energy balance for the vapor space is as follows

$$dM_{V(i)} Cp_{V(i)} \frac{dT_{V(i)}}{dt} = m_{bg(i)} Cp_{bg(i)} T_{bg(i)} + m_{pg(i)} Cp_p - U_{(i)} A_{C(i)} \Delta T_{(i)} \quad (8)$$

Where $U_{(i)}$ is the heat transfer coefficient, $A_{C(i)}$ is the heat transfer area and $\Delta T_{(i)}$ is the log mean temperature difference (LMTD), given by

$$\Delta T_{(i)} = \frac{\left(T_{c(i)} - T_{wa(i)}\right) - \left(T_{c(i)} - T_{wa(i+1)}\right)}{Ln\left[\frac{\left(T_{c(i)} - T_{wa(i)}\right)}{\left(T_{c(i)} - T_{wa(i+1)}\right)}\right]} \quad (9)$$

The mass balance for the condenser tubes can be written as

$$m_{wa(i)} = m_{wa(i+1)} \quad (10)$$

The salt balance of circulating brine is given by

$$X_{wa(i)} = X_{wa(i+1)} \quad (11)$$

It is assumed that mass of brine in the condenser tubes remains constant and there is no accumulation of salt in the condenser tubes. Thus,

$$\frac{dM_{WA(i)}}{dt} = 0 \quad \text{and} \quad M_{WA(i)} \frac{dX_{WA(i)}}{dt} = 0 \quad (12)$$

The energy balance for the condenser tubes yields

$$M_{WA(i)} Cp_{ba(i)} \frac{dT_{WA(i)}}{dt} = m_{wa(i+1)} Cp_{ba(i+1)} T_{wa(i+1)} - m_{wa(i)} Cp_{ba(i)} T_{wa(i)} + U_{(i)} A_{C(i)} \Delta T_{(i)} \quad (13)$$

The saturation temperature of the vapor above the flashing brine will be less than the brine temperature by the boiling point elevation (BPE), non-equilibrium allowance (NEA) and loss in saturation temperature due to other pressure losses ΔT_P,

$$T_{V(i)} = T_{b(i)} - BPE_{(i)} - NEA_{(i)} - \Delta T_{P(i)} \quad (14)$$

3.1.2. Last stage Model

The brine mass conservation equation is

$$\frac{dM_{B(n)}}{dt} = m_{b(n-1)} + m_k - m_{wa(n)} - m_{bd} - m_{bg(n)} \quad (15)$$

The balance of dissolved salt is

$$M_{B(n)} \frac{dX_{B(n)}}{dt} = m_{b(n-1)} X_{b(n-1)} + m_k X_k - m_{wa(n)} X_{w(n)} - m_{bd} X_{bd} \quad (16)$$

The energy balance for the brine is,

$$M_{B(n)}Cp_{bd}\frac{dT_{B(n)}}{dt} = m_{b(n-1)}Cp_{b(n-1)}T_{b(n-1)} + m_k Cp_k T_k - m_{wa(n)}Cp_{wa(n)}T_{wa(n)} - Cp_{bd}m_{bd}T_{bd} - m_{bg(n)}h_{bg(n)} \quad (17)$$

The conservation equations for distillate tray vapor space and the condenser tube can be written in a similar manner as that for the repeated stage (i), Eqs. (5) to (14).

The formulated basic equations include some physical properties which are correlated with subsidiary equations as functions of these stage variables, it can be found in[17].

3.1.3. Modeling of the Brine Heater

The brine heater is a shell and tube type heat exchanger in which, heating steam condenses outside the tube surface, exchanging its latent heat with the brine. The brine flows through the tubes and consumes the energy supplied by the steam. A schematic of the brine heater is shown in Figure 3.

The brine heater (BH) has a steam flow, m_S coming in at temperature T_S, and the condensate, m_L going out through at temperature T_L. The incoming steam is considered to be saturated and at the temperature of condensate in the sump as in the real plant

The mass and energy balance equations are given by

$$M_{BH}Cp_{BH}\frac{dT_{BH}}{dt} = m_S(h_S - h_L) - m_{wa(1)}(Cp_{wa(0)}T_{wa(0)} - Cp_{wa(1)}T_{wa(1)}) \quad (18)$$

$$M_{BH} = \rho_{BH}V_{BH} \quad (19)$$

$$m_{wa(1)} = m_{wa(0)} \quad (20)$$

$$m_L = m_S \quad and \quad T_L = T_S \quad (21)$$

$$X_{wa(1)} = X_{wa(0)} \quad (22)$$

$$V_{BH} = A_{BH}LT_{BH}NT_{BH} \quad (23)$$

Where, (0) refers to the 0^{th} stage i.e. the brine heater (BH).

4. Solution Procedure

The resulting sets of governing equations are solved numerically by using Runge-Kutta method. A computer package based on Mat-lab with menu driven user interface, interactive graphic interface, long double precision, mouse support, and printer support is implemented, the main output windows are shown in appendix (A). Starting from the initial state, the simulation strategy involves the procedure carried out in the following steps:

1. The algebraic variables are determined from correlations and algebraic equations.
2. The right hade side (R.H.S) of all the ODE's is calculated by using the algebraic variables determined in step 1.
3. A numerical solution using Runge-Kutta method uses the R.H.S to calculate the new values of all the differential variables for the next simulation time till a certain convergence.
4. Increment time by Δt and repeat steps 1 to 5, till a steady state is approached

5. Model Validation

The validity of the present model was checked by comparing the steady state results with some of the previous results and the actual data for Kuwait, Benghazi and Zuitina MSF plants. The comparisons are shown in Tables (1) to (3). Table (1) shows the comparison between the present predictions and actual data of Kuwait for the top brine temperature (TBT), the recirculating brine temperature that enters the brine heater (TW(1)), the product flow rate (mpr), the product temperature (TPr), the blow-down temperature and the performance ratio, which is defined as the mass flow rate ratio of distillate product to the heating steam. Table 2 shows the comparison between the present prediction of the top brine temperature (TBT), recirculating brine temperature enters the brine heater (TW(1)), product flow rate (mpr), product temperature (TPr), blow-down temperature and performance ratio with the actual data Zuitina MSF plant. The comparison shows good agreement between the present predictions and the actual data. Therefore, the present model is considerably valid to accurately predict the performance characteristics of MSF desalination plant at both steady states and transient.

6. Case Study

In this case study, the transient performance characteristics have been predicted utilizing the design data of Zuitina MSF desalination plant. The plant is constructed with a cross-tube-type multi-stage flash (MSF) evaporator with recirculating brine, and multi-stage condensers with two sections. The plant is composed of a heat recovery section (18 stages) and a heat rejection section (3 stages). Some of the design data of the Zuitina plant is listed in Table 3.

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Table 1. Comparison between the present predictions, and some pervious data

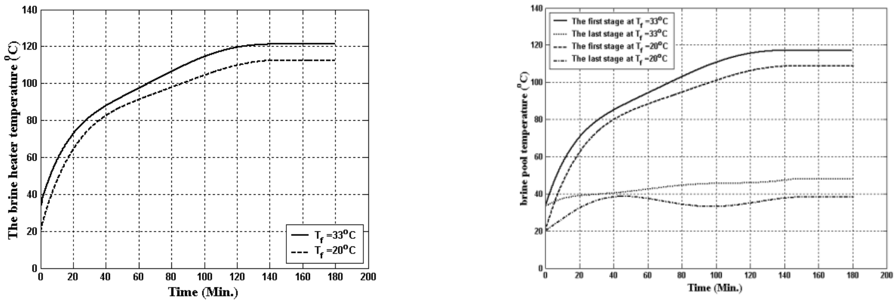
Operating Parameters		Kuwait MSF plant	Al-Shayji [19]	Present predictions	(%) Deviation
Input	No. of Stages	24	24	24	---
	m_f , (T/h)	9629.3	9629.3	9629.3	---
	T_f , (^oC)	32.22	32.22	32.22	---
	m_S, (T/h)	140.862	140.862	140.862	---
	T_S , (^oC)	100	100	100	---
	X_F , ppm	45,000	45,000	45,000	---
Output	TBT , (^oC)	90.56	89.1	90.672	0.12
	$T_W(1)$, (^oC)	83.20	82.22	82.429	0.93
	m_{pr} , (T/h)	1127.7	1159.8	1121.5	0.55
	T_{Pr} , (^oC)	38.60	36.54	39.047	1.15
	T_{bd} , (^oC)	40.5	38.44	39.521	2.42
	PR	8.005	7.76	7.961	0.48

Table 2. Comparison between the present predictions and the data from Zuitina MSF plant

Operating Parameters		Zuitina MSF plant	Present predictions	(%) Deviation
Input	No. of Stages	21	21	---
	mf , (T/h)	2800	2800	---
	Tf, , (^oC)	28	28	---
	mS , (T/h)	52	52	---
	TS , (^oC)	124	124	---
	XF , ppm	35,000	35,000	---
Output	TBT , (^oC)	118	117.8612	0.12
	TW(1) , (^oC)	109.3	109.4263	0.11
	mpr , (T/h)	416.7	413.6298	0.74
	TPr , (^oC)	37	39.3282	5.9
	Tb(N) , (^oC)	38.2	40.3367	5.6
	PR	8.0128	7.9544	0.73

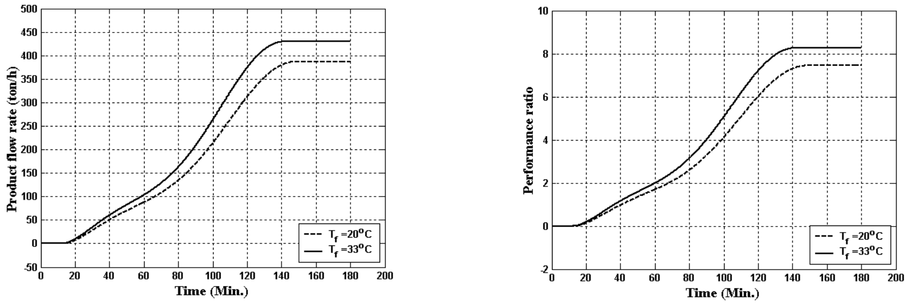
Table 3. The condenser tube bundle specifications

Parameter	Recovery section	Rejection section	Brine heater
Tube material	CuNi10Fe	Titanium	CuNi30Fe
No. of tubes per stage	1313	1410	1302
Tube length (m)	9	9.4	13.09
Outside diameter (mm)	24	19	24
Wall thickness (mm)	0.9	0.5	1
Heat transfer area (m^2)	891	791	1274
Fouling factor (m^2.K/W)	0.00009	0.000178	0.000178



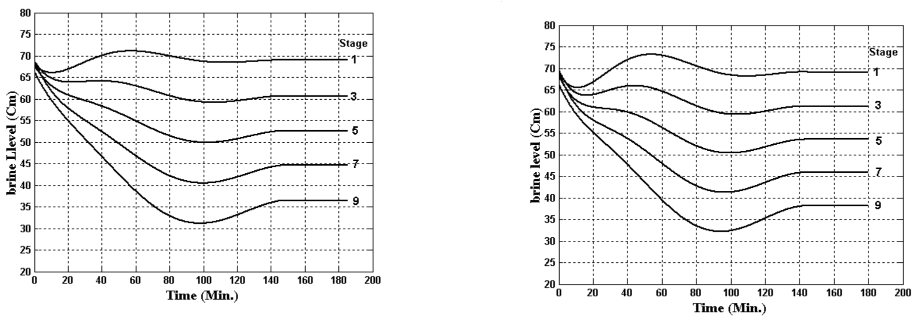
(a) At brine heater outlet (b) At the outlet of the first and the last stages

Figure 8. The time dependent flashing temperature at different seawater inlet temperatures



(a) Product flow rate (b) Performance ratio

Figure 9. Effect of seawater inlet temperature on product flow rate and performance ratio



(a) At T_f =20 °C (b) At T_f =33 °C

Figure 10. Brine levels at start up period for Zuitina plant at two seawater inlet temperatures

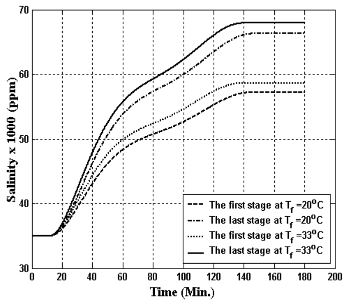


Figure 11. Stage wise salinity at two different seawater inlet temperatures

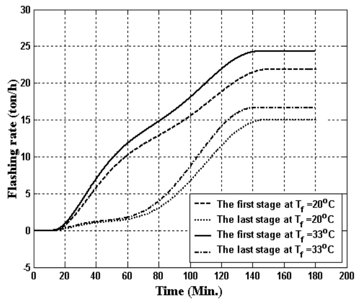


Figure 12. Stage wise flash rate at two different seawater inlet temperatures

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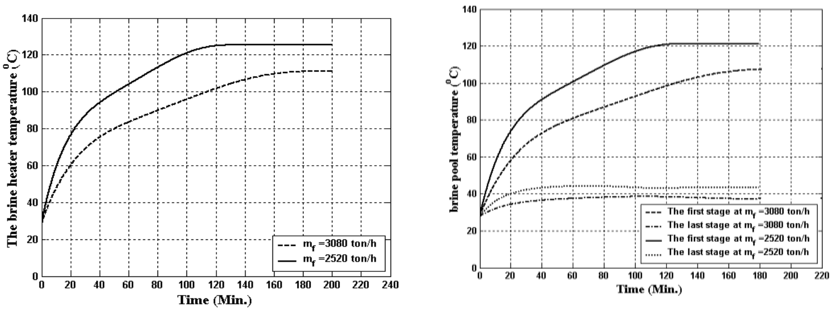
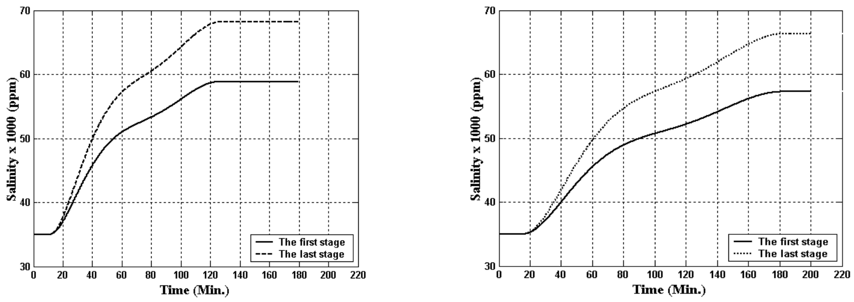


(a) Brine heater outlet temperature (b) Brine pool temperature

Figure 13. Brine temperatures at different seawater inlet flow rate



(a) m_f=2520 ton/h (-10% of design point) (b) m_f =3080 ton/h (+10% of design point)

Fig.ure 14. Stage wise salinity at two different seawater inlet flow rate

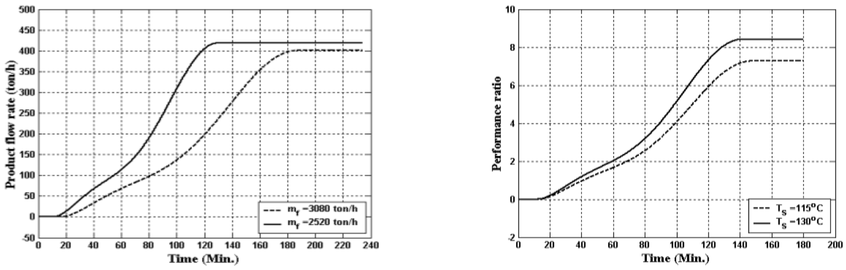
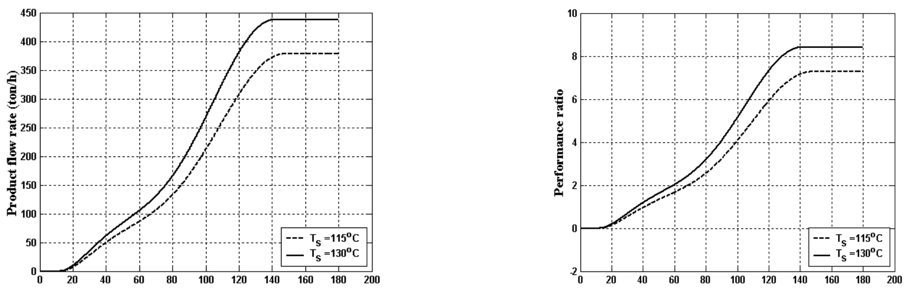


Fig. 15. Effect of seawater flow rate on the product flow rate and performance ratio (±10% of design point)

i. Effect of the Heating Steam Temperature

One of the most important parameter that affects the MSF desalination process is the temperature of heating steam. The top brine temperature is dependant on the heating steam temperature which affects the overall plant performance characteristics as shown in Figure 16. The plant performance enhances with the increase in the heating steam temperature. This is due to the fact that as heating steam temperature increases the top brine temperature increases which enhances the flashing rate and product flow rate and in accordance enhances the plant performance ratio. An increase in the flashing rate results in the decrease of the brine levels in the first stage and subsequently to the other stages (as shown in Figure 17). Unfortunately, the extra increase in the heating steam temperature may increase the scales due to the increase in both salinity and the overall temperature level in the plant. Therefore, the heating steam temperature is limited to about 130 ^{o}C in most of the commercial MSF plant



(a) Product flow rate (b) Performance ratio

Figure 16. The effect of heating steam temperature on transient product flow and performance ratio

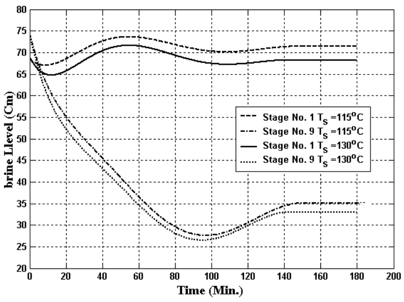


Figure 17. Brine level at two different heating steam temperatures

7. Conclusion

This work has presented a dynamic simulator for the analysis of MSF desalination plants using a dynamic analysis. The developed model is based on the basic equations of mass, momentum and energy. The proposed was able to investigate the affect of some key parameters such as seawater concentration and other thermal parameters that may affect the general performance of the MSF plant during transient as well as steady state operation condition. The proposed model was validated by using data from previous theoretical studies as well as actual data obtained from an operating MSF plant. The validation results showed good agreement with the predicted and the real case measured values. Based on data derived from a number of selected runs, the following conclusions may be drawn:

1. The seawater inlet flow rate (mf) had a very strong effect on the system performance, where its decrease results in increasing the system temperature and the flashing rate and subsequently the performance ratio. This is the opposite effect to increasing the seawater inlet flow rate.
2. An increase in the heating steam temperature (T_S) results in an increase of the system temperature, which improves the flashing efficiency as well as the net product flow rate. Also, this results in a decrease of the brine level that might result in vapor leaking across the stages. This is the opposite effect to decreasing the heating steam temperature.
3. Sea water temperature (T_f) has a considerable affect on the product of the plant as well as the performance ratio, and it has a slight effect on the top brine temperature and brine levels.

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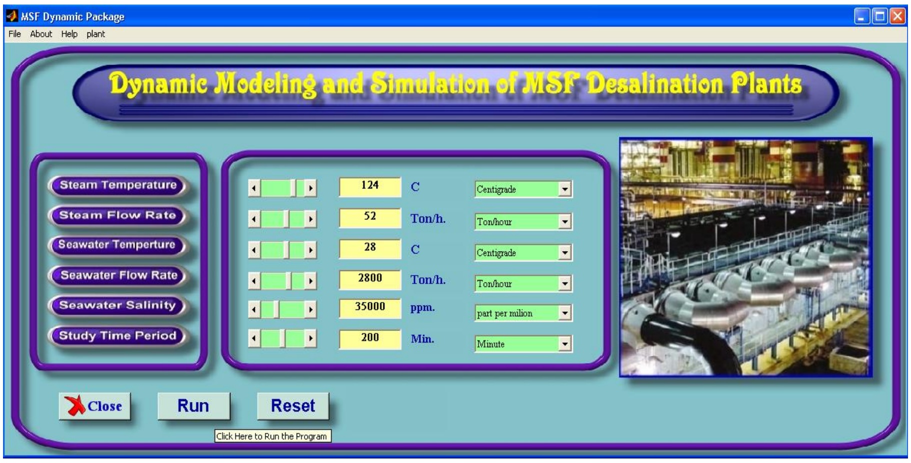
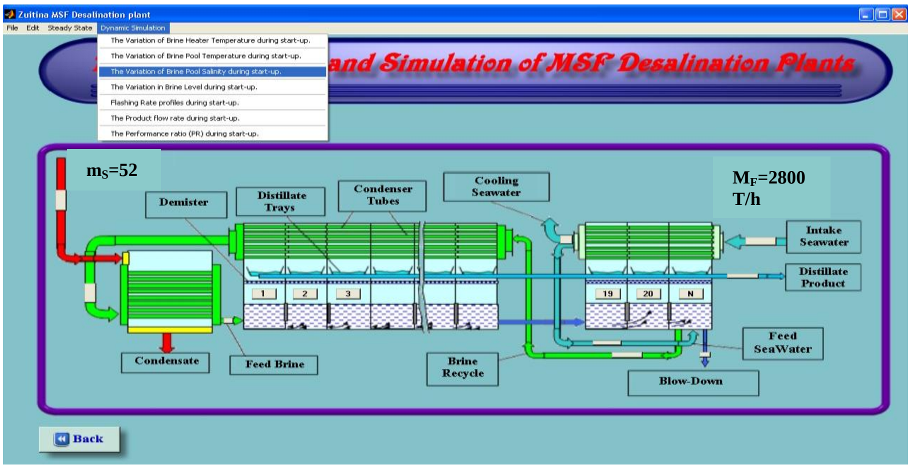
Appendix



The Mat-lab graphical user interface for the interactive simulation



The main out-put screen